UNITED STATES PATENT APPLICATION FOR

PROCESS AND APPARATUS FOR
CONVERTING OIL SHALE OR TAR SANDS TO OIL

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PROCESS AND APPARATUS FOR
CONVERTING OIL SHALE OR TAR SANDS TO OIL

FIELD OF THE INVENTION

The present invention relates to a continuous process for producing synthetic crude oil (SCO) from oil shale or tar sand and an apparatus for its practice. More specifically, the present invention provides a process for treating dry tar sand or shale without prior beneficiation, in a

5   reactor operating at elevated pressure and temperature conditions, in the presence of substantially only hydrogen gas.

RELATED APPLICATIONS:

This application is a continuation-in-part of application Serial No 09/058,184 filed on

10   April 10, 1998, which is a continuation-in-part of application Serial No. 08/843,178, filed on April 14, 1997, now U.S. Patent No. 5,902,554, which in turn is a division of application Serial No. 08/551,019, filed October 31, 1995, now U.S. Patent No. 5,681,452, each of which is hereby incorporated herein by reference.

15   ## BACKGROUND OF THE INVENTION

There are some tar sand systems that are successful in making synthetic crude oil (SCO), such as those in the Canadian Athabasca tar sand area that surface mine and process the tar sands, where they first separate sand (85%) from bitumen (15%) to avoid processing the sand in the reaction systems. The separated bitumen is converted to sweet, light crude oil by conventional

20   refinery type operation. Separation of the sand from the bitumen requires beneficiating operations such as floatation cells and secondary separation equipment and processing and

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equipment to prepare the tar sand for flotation. In these systems, tailing oil recovery is necessary to clear the sand for disposal, however the sand is not completely cleared of bitumen.

Existing technology uses a large number of physical and chemical processing units for the treatment of wet tar sands, e.g., fluid cokers, LC finer, tumblers (being phased out by hydro-

5   pumping), beneficiators including: primary separation vessels with floatation cells and secondary separation systems necessary to recover the bitumen from the tar sand; tailing oil recovery systems which result from the sand not being completely cleared of bitumen; tailing settling ponds which are necessary to settle and separate fine clays and other undesirable solids from the water required for floatation since the water must be reused to maximize clean-up to reduce

10   environmental problems. These systems require large facilities along with the maintenance and reclamation required.

For example, U.S. Patent Nos. 5,340,467 and 5,316,467 to Gregoli, et al. relate to the recovery of hydrocarbons (bitumen) from tar sands. In the Gregoli, et al. patent process, tar sand is slurried with water and a chemical additive and then the slurry is sent to a separation system.

15   The bitumen recovery from tar sand processes described in U.S. Patent Nos. 5,143,598 to Graham et al. and 4,474,616 to Smith, et al. also involve the formation of aqueous slurries. Other processes involving slurries, digestion, or extraction processes are taught in U.S. Patent Nos. 4,098,674 to Rammler, et al., 4,036,732 to Irani, et al., 4,409,090 to Hanson, et al., 4,456,536 to Lorenz, et al. and Miller, et al.

20   In situ processing of tar sand is also known as seen from the teachings of U.S. Patent Nos. 4,140,179, 4,301,865 and 4,457,365 to Kasevich, et al. and 3,680,634 to Peacock, et al.

U.S. Patent No. 4,094,767 to Gifford relates to fluidized bed retorting of tar sands. In the process disclosed by the Gifford patent, raw tar sand is treated in a fluidized bed reactor in the presence of a reducing environment, steam, recycle gases and combustion gases. The conversion

25   of the bitumen, according to the Gifford patent, is through vaporization and cracking, thereby

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leaving a coked sand product. The steam and oxygen, according to Gifford are "injected into the fluidized bed in the decoking area above the spent sand cooling zone, and below the input area in the cracking zone for fresh tar sand."

The process and apparatus of the present invention avoid the use of the large number of

5   physical and chemical processing units used in the processing of wet tar sand by using a single continuous reactor system to hydrocrack and hydrogenate the dry tar sand. Moreover, because the present invention directly hydrogenates dry tar sand, larger quantities of valuable sweet, light crude oil are obtained. Moreover, with the present invention, less gas and substantially no coke is produced.

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BRIEF SUMMARY OF THE INVENTION

The present invention relates to a continuous process for converting oil bearing material, e.g., oil shale or tar sand, and an apparatus for its practice.

Accordingly, one aspect of the present invention is to provide a continuous process and

15   an apparatus for its practice where oil bearing material such as the kerogen in oil shale or the bitumen in tar sand is continuously treated.

Another aspect of the present invention relates to the treatment of dry tar sand.

An object of the present invention is providing a method and apparatus for converting a tar sand or shale feed to oil which can be conducted in the absence of a benificiation processes

20   such as, for example, a hot-water extraction process to separate sand or other non reacting solids from bituminous or oil-bearing material in the feed.

An object of the present invention is providing a process for converting tar sand to oil through the use of substantially only hydrogen.

Another object of the present invention is providing a heat recovery process whereby

25   hydrogen provides the heat necessary to bring the raw tar sand up to reactor temperature.

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A still further object of the present invention is providing a process where hydrogen is used for hydrocracking and hydrogenating the bitumen in the tar sand or oil shale.

A further objective of the present invention is providing a process for using recycle and make-up hydrogen as a heat transfer vehicle.

5   A still further object of the present invention is providing an improved process for producing oil from tar sand or shale by reacting the tar sand or shale with substantially only hydrogen in a fluidized bed reactor, wherein the fluidizing medium is substantially hydrogen.

Yet another object of the present invention is providing a fluidized bed process where one inch or less size tar sand or shale pieces are fed into a fluidized bed reactor near the bottom of the

10   reactor and spent sand and reaction products are removed from near the top of the reactor.

Still another object of the present invention is providing a method of recycling unreacted hydrogen that exits a reactor in which tar sand or oil shale is converted to oil. The method includes purging impurities in the exiting recycle hydrogen stream by pressure swing adsorption, maintaining the hydrogen at more than about 450 psi throughout the recycle process, admixing

15   fresh hydrogen to the recycle hydrogen stream to form a mixture, and feeding the mixture into the reactor.

These objectives can be achieved by providing a process for producing oil from an oil bearing feed such as tar sand or oil shale. The process comprises introducing the feed in a fluidizable form into a fluidized bed reactor. A fluidizing medium enters the fluidized bed reactor

20   where it contacts and fluidizes the fluidizable feed. The fluidizing medium includes at least hydrogen. The fluidized feed forms a fluidized bed where the feed reacts with substantially only the hydrogen at a temperature of at least 900°F. The reaction products include synthetic crude oil and spent solids which are discharged from the fluidized bed reactor.

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BRIEF DESCRIPTION OF THE DRAWINGS

FIG. 1 shows the flow diagram of one embodiment according to the present invention.

FIG. 2 shows a fluidized bed reactor for converting bitumen in tar sand to viable products in accordance with the present invention.

5   FIG. 3 shows a stand-alone fired heater used in the process according to the present invention.

FIG. 4 shows a compressor for supplying the hydrogen for use in the present invention.

FIG. 5 shows the flow chart of an acid gas recovery system for use in the present invention.

10   FIG. 6 shows the mass balance for one embodiment of the present invention.

FIG. 7 shows a flow diagram of a second embodiment according to the present invention.

FIG. 8 shows a fluidized bed reactor and lock hoppers of the second embodiment according to the present invention.

In FIGS. 1-6, common elements are similarly identified except for the "figure number" designation.

15   Thus, all elements depicted in FIG. 1, start off with the number 1, e.g., the reactor in FIG. 1 is identified as "104" and in FIG. 2 the same reactor is identified as "204."

DETAILED DESCRIPTION OF THE INVENTION

In the present invention the hydrocarbon content of the hydrocarbon bearing solids, e.g.,

20   dry tar sand or oil shale is reacted in a fluidized bed reactor with hydrogen and the process is operated to avoid decompression of the hydrogen. In the present invention, the hydrocarbon bearing solid does not include bituminous or anthracite coals or similar type material. A first portion of a substantially only hydrogen stream is used to feed the oil shale or tar sand, which has been comminuted and reduced in size to form particles that are capable of being fluidized, e.g.,

25   fluidizable, into the reactor. A second portion of the hydrogen stream is used as the fluidizing

medium. The hydrogen stream that is used in the present invention is formed from fresh make-up hydrogen and recycle hydrogen generated during the process, or obtained from other hydrogen producing processes. A mixed fresh-make-up and recycle hydrogen stream is discharged from a compressor at a first temperature and pressure, and a portion is diverted for admixture with the

5   fluidizable particles of tar sand or oil shale which are injected into the fluidized bed reactor in a fan like flow, at an acute angle relative to the vertical axis of the reactor or a horizontal plane. The remainder of the hydrogen stream at said first temperature is indirectly heated to a second higher temperature by indirect heat exchange with overhead products from the fluidized bed reactor. The hydrogen stream at said second temperature is conveyed to a direct fired heater

10   where the hydrogen stream is heated to a third temperature higher than said second temperature and then used as the fluidizing medium in the reactor to fluidize the tar sand or oil shale fluidizable particles that have been injected with the first portion of the hydrogen stream.

In the fluidized bed reactor the bitumen in the tar sand, or the kerogen in the oil shale, and hydrogen are reacted via endothermic and exothermic reactions to produce spent tar sand or

15   oil shale and an overhead product stream that contains hydrogen, hydrogen sulfide, sulfur gases, $C_1 + C_2$ hydrocarbons, ammonia, fines (sand particles and clay) and vaporous products. The overhead product stream is first separated in cyclone separators within the reactor which help maintain the bed level and separate solids. The first separated overhead product is conveyed to a series of additional separators to provide a substantially particle free clean product stream. The

20   cleaned product stream at a first temperature is conveyed to a first heat exchange unit where heat is transferred to a second portion of the hydrogen stream and results in a product stream at a second temperature lower than said first product stream temperature. The product stream at said second temperature is conveyed to a condenser to further reduce its temperature to a third temperature lower than the second product stream temperature. The product stream at said third

25   temperature contains liquid and gas fractions and is conveyed to a separator where the gas

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fraction is removed, sent to an amine scrubber, and recycled as a scrubbed recycle hydrogen stream, and the liquid fraction is removed as oil product (SCO). The cooled, absorbed overhead hydrogen stream is conveyed to a heat exchanger where it contacts spent tar sand or spent shale and its temperature is elevated due to heat transferred from the spent discharge. The hydrogen

5   stream at the elevated temperature is conveyed to a cyclone separator, or other suitable separating devices to remove particles. It then flows to the amine system to regenerate the amine solution. It is eventually conveyed to a compressor where it is combined with fresh make-up hydrogen for use in the fluidized bed reactor as the first and second hydrogen stream portions.

The invention will now be described with reference to the figures. **FIG. 1** is a flow chart

10 of one embodiment of the present invention where tar sand is converted to oil. In accordance with the present invention, tar sand from the run of mine conveyor belt 101 is continuously fed to any suitable sizing equipment 102 for classifying tar sand, at a temperature of about 50°F. Tar sand is composed of bitumen and sand.

The bitumen in the tar sand that is processed in the present invention normally contains

15 heavy metals which catalytically help promote the endothermic and exothermic reactions in reactor 104. However, it may be advantageous to add additional catalyst. The tar sand processed in accordance with the present invention is exemplified by the following, non-limiting example:

TAR SAND FEED

sand		84.6 wt. %
bitumen		15.4 wt. %
carbon	83.1 wt. %	
hydrogen	10.6 wt. %	
sulfur	4.8 wt. %	
nitrogen	0.4 wt. %	
oxygen	1.1 wt. %	
nickel	75 PPM	
vanadium	200 PPM	
	100 wt. %	100 wt. %

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In the present invention dry tar sand having an average particle size of that of sand is conveyed through conduit 103 as the feed for fluidized bed reactor 104, discussed in greater detail in **FIG. 2**. Tar sand particles which are oversized are either recycled to the sizing equipment 102, or conveyed to any suitable equipment for reducing the size of the oversized feed. In the present

5 invention, the phrase "dry tar sand" means, under atmospheric conditions, a friable, non-sticky, easily handled, substantially free flowing material.

Tar sand is fed through pressure feeder rotary valves 104A which are circumferentially positioned adjacent and around the upper end of the fluidized bed reactor 104, which is described in detail greater in **FIG. 2**. The rotary feeders 104A are positioned at an angle of between 20 and

10 60 degrees relative to the vertical reactor axis in order to "fan feed" the fluidizable sized tar sand into the top of the reactor 104. More uniform dispersion of the tar sand in the fluidized bed reactor can be obtained when three or more rotary feed valves 104A are positioned equidistantly around the circumference of the reactor. Although three feeders 104A are preferred, the size of the reactor and the degree of fanning desired will control the number of valve feeders. Thus,

15 there could be 4, 5, 6, 7 or more valve feeders used in the present invention.

High pressure hydrogen is conveyed through lines 138 to the feeders 104A, at a pressure of between 625 psi and 700 psi, preferably about 635 psi, to assist in injecting, feeding and dispersing the tar sand into reactor 104.

The process performed in fluidized bed reaction 104 involves hydrocracking, which is an

20 endothermic reaction, and hydrogenation, which is an exothermic reaction, which reactions are conducted to favor the production of liquid fuels and minimize the production of gas yields. The reactor operates at temperatures of between 800°F and 900°F, preferably closer to 800°F to avoid cracking the large fragments of hydrogenated bitumen in the tar sand.

It is advantageous to conduct the endothermic hydrocracking and exothermic

25 hydrogenating processing of tar sand in reactor 104 in a predominantly hydrogen gas

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environment. The hydrogen atmosphere in reactor 104 is maintained at about 600 psi by fresh

make-up hydrogen conveyed through line 130 from a hydrogen plant and a hydrogen recycle

stream 129 which contains cleaned-up hydrogen. The volume of recycle hydrogen to fresh make-

up hydrogen is preferably at least about 26 to 1.

5   Advantageously all the high pressure hydrogen for the process of the present invention,

for reaction in reactor 104 and the various heat exchange operations, is provided by the steam

powered compressor 132. Compressor 132 receives fresh make-up hydrogen which is conveyed

through line 130 and recycle hydrogen which is conveyed through lines 129, 140, 142, 144 and

131. Compressor 132 is powered by steam conveyed through line 162 from direct fired heater

10   135.

Reactor 104 operates in a highly agitated fashion insuring almost instant and complete

reaction between the bitumen components and hydrogen. The residence or retention time of the

tar sand in reactor 104 is about 15 minutes, but could be between 10 and 20 minutes, depending

on the throughput and efficiency of the reactor process. The pressure drop from the bottom to the

15   top of the reactor 104 is about 35 psi.

Overhead products from reactor 104 are discharged from reactor 104 through cyclone

separators 104C, while solids are discharged through separator section 104B located at the lower

end of reactor 104. The cyclones separators 104C discharge an overhead stream, e.g., gas and

vapor reaction components, off-gas and product, through their upper ends into line 110, while

20   separated solids are discharged through the lower ends of the dip legs. The cyclone separators

104C extend about 20 feet down into the reactor 104 and establish the bed height in the reactor

104.

The hot spent tar sand is continuously discharged at a pressure of about 635 psi and a

temperature of about 800°F through lock hopper valving arrangement 104B in the lower end of

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reactor 104 into line 105 which conveys the discharged material to spent sand heat exchangers 106 and 108.

The reactor overhead stream from the cyclone separators 104C is discharged into line 110, at a temperature of about 800°F and a pressure of about 600 psi. The overhead stream

5 discharged from the reactor 104 still contains dust and dry waste particles, and is first conveyed through line 110 to cyclone separator 111 where solids are separated and removed through line 150. The gaseous effluent from separator 111 is conveyed through line 112 to an electrostatic precipitator 113 for the final cleanup. The cleaned overhead stream from precipitator 113 is removed and conveyed through line 114, and separated solids are discharged through line 151.

10 Cyclone separator 111 and electrostatic precipitator 113 are of conventional design and one of ordinary skill in the art practicing the present invention can select suitable devices for performing the described operation.

The cleaned stream from the precipitator 113, product, vaporous components, and off gas, are conveyed to in-and-out heat exchanger 115 through line 114. In the in-and-out

15 exchanger 115 the cleaned stream from line 114 is brought into indirect heat exchange relationship with hydrogen being conveyed through line 133, from compressor 132, i.e., recycle and fresh make-up hydrogen, whereby heat is transferred from the cleaned stream to the hydrogen in line 133 prior to the hydrogen stream entering the fired heater 104. The cooled and cleaned stream, products, vaporous components, off-gases, from heat exchanger 115 is discharged into

20 line 116 while hydrogen is discharged into line 134 which conveys the hydrogen to the direct fired heater 134.

The cooled stream being conveyed through line 116 is introduced into condenser 117 and is discharged at a temperature of about 100 °F into line 118. The vapor and gas stream from the condenser is conveyed through line 118 at a temperature of 100°F and is introduced into separator

25 119 where vapors and liquid are separated and discharged.

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Since the gas stream has been cooled down to about 100°F and is still at a pressure of 480 psi, all carbon compounds C_3 and above have been condensed are removed from the separator 119 through flow line 155 to storage. Sour water from the separator is discharged through flow line 154. The crude oil product stream in line 155 is a mixture of naphtha and gas oils having an

5   A.P.I. of approximately 33.5 and is a light sweet crude. The gas stream in line 120 is conveyed to a scrubbing system, e.g., at least one amine absorption column 121 where sulfur components, e.g., hydrogen sulfide and sulfur dioxide gases, are absorbed and discharged through line 122 and conveyed to a suitable sulfur recovery plant. The amine absorption system 121 is described in greater detail in **FIG. 5**.

10   The only gases not absorbed and removed in absorption system 121 are unreacted recycle hydrogen and $C_1 + C_2$ hydrocarbons which are conveyed through line 129 to heat exchangers 106 so that the spent tar sand is cooled and the recycle hydrogen and $C_1 + C_2$ hydrocarbons is heated and discharged into line 140. The C_1 and C_2 hydrocarbons in line 129 will not be absorbed nor condensed but will be recycled with the unreacted hydrogen after processing in units 141, 143

15   and 145 discussed hereinafter. The C_1 and C_2 hydrocarbons will reach equilibrium within the reactor 104 at about 2% and will then add to the production of crude oil per ton of tar sand. A small offset will be the increase in the recycle stream.

As discussed above, the spent sand from the reactor 104 is discharged into a succession of heat exchangers 106 and 108. The first heat exchanger 106 cools the sand from 792°F to

20   400°F using cool recycle hydrogen being conveyed through line 129. The cooled spent sand is conveyed in line 107 from heat exchanger 106 and introduced into a second heat exchanger 108 so that the sand is cooled by cold air introduced through line 180 from blower 181 and through line 182, before discharging. The air heated by the spent sand is discharged into line 183 which conveys the heated air to fired heater 135 for combustion therein. Although two heat exchangers

25   are shown, the invention contemplates using more if necessary.

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The heated and partial recycle hydrogen stream conveyed through line 140 is introduced into cyclone 141, discharged into line 142 which conveys the stream to precipitator 143, and then through line 144 for introduction into exchanger 145.

5   **FLUIDIZED BED REACTOR**

FIG. 2 schematically shows the pressurized, continuously operating fluid bed reactor 204 in accordance with the present invention. Sized and screened tar sand or shale are conveyed through lines 203 and fed through pressure feeder rotary valves 204A into the top of the reactor 204. A portion of the gases processed in compressor 132 (**FIG.** 1), and heated in fired heater 135

10   (**FIG.** 1) are conveyed by line 236 and introduced into fluidized bed reactor 204 in an upward direction to fluidize the bed of the reactor 204. Another portion of the hydrogen gas from line 133 is conveyed through line 237 to tar sand feed valves 204A through lines 238. Another portion of the hydrogen gas feed from line 237 is diverted through lines 239 and injected into the separator section 204B, at the bottom end of reactor 204. Hydrogen conveyed in lines 239 is

15   injected into the separator section 204B of reactor 204 through injectors which are located at the ends of flow lines 239 (not shown) and aid in heat retention in the reactor system and spent sand discharge through line 205.

High temperature and high pressure hydrogen (make-up and recycle) after passing through the direct fired heater 135, is introduced into reactor 204 from line 236. Reaction

20   products and unreacted hydrogen exit the reactor through internal cyclones 204C ensuring even flow out of the reactor. Although two cyclone separators are shown, the invention contemplates using as many as necessary to provide even flow of product gases from reactor 204 and bed height maintenance. The hot reactor effluent stream in line 210 is then conveyed to physical and chemical units, described in **FIG.** 1 for cleanup heat recovery and product separation.

25   **DIRECT FIRED HEATER**

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As discussed above with reference to **FIG.** 1, a portion of the fresh make-up and cleaned recycle hydrogen from the compressor is conveyed to a direct fired heater. **FIG.** 3 schematically shows a fired heater 335 (135) that is designed to balance out the total energy required to operate the reactor system. Preheated air conveyed through feed lines 383 (183) is combusted with fuel in the radiant section of fired heater 335 (135) and elevates the temperature of the recycle and make-up hydrogen that is conveyed through line 334 (134). The fuel that is combusted is obtained from the C_3 fraction, e.g. propane, or natural gas produced or purchased from the described process or other sources. The hydrogen stream in lines 334 (134) has been preheated in the reactor in-out exchanger 115 to approximately 750°F. Since the hydrogen stream is circulated through the radiant section of the heater 335 the temperature of the hydrogen stream is elevated to a temperature of about 1200°F. Circulation of the hydrogen stream through line 133, 134, exchanger 115 and fired heater 335 is maintained by compressor 132 so that the 1200°F hydrogen stream can be introduced into reactor 104 (**FIG.** 1) or 204 (**FIG.** 2).

Waste heat from the radiant section of direct fired heater 335 is recovered in convection section 335A (135A), 335B (135B) and 335C (135C). Steam separated in drum 360 (160) is discharged into line 361 (161) and introduced into convection section 335A (135A) where the steam temperature is raised from about 596°F to about 800°F. After passing through convection section 335A (135A), the super heated, high pressure steam is conveyed through line 362 (162) to drive the steam turbine 163. Reduced temperature and pressure steam from turbine 163 is conveyed to steam condenser 165 and the condensate recirculated via line 166 and pump 166A. The flow from pump 166A is conveyed through line 168 (368) and combined with make-up water from line 167. The water being conveyed in line 268 is introduced into convection section 335C (135C), heated and discharged through line 369 (169) for further processing, e.g., deaeration.

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Steam drum 360 (160) separates steam which is conveyed to radiant section 335A (135A) through line 161 to produce superheated steam for the turbine compressor 163.

The steam circulation loop include steam drum 360 (160), line 370 (170), recirculation pump 371 (171) and lines 372-373 (172-173) which conveys boiler water through radiant section

5 335B (135B) and back into drum 360 (160). Water for the boiler system is provided through feed line 467 (167) which flows into line 468. Line 468 is similar to flow line 168, 368 which communication with line 169 through connection section 335a (135a) to discharge.

As discussed above, convection section 335A (135A) super heats steam which is conveyed through line 362 (162) to drive compressor turbine 163, which drives compressor 132.

10 Steam is generated in convection section 335B (135B) and make-up water and turbine condensate for boiler feed water are preheated in convection section 335C (135C).

COMPRESSOR SYSTEM

FIG. 4, schematically shows a compressor 432 (132) driven by a high pressure steam

15 turbine 463 (163) required to maintain circulation of gases to operate the reactor system 104. Make-up hydrogen 430 (130) and recycle hydrogen 431 (131), at approximately 450 psig and 100°F are pressurized by the compressor 432 (132) to approximately 670 psig and 122°F and discharged into line 133 which conveys and introduces the high pressure hydrogen into the in-out exchanger 115 to be further heated by exchange with reactor product gases.

20 High pressure steam in line 162, 362, at 1500 psig and 800°F drives the turbine 463 (163). Exhaust steam 464 (164) is condensed in condenser 465 (165), and along with make-up water 467 (167) is fed to the fired heater convection section 135C, 335C for preheating and reuse as boiler feed water make-up.

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PRODUCT SEPARATION

The product separation of **FIG.** 1, components will be described in greater detail with reference to **FIG.** 5, which schematically shows the product separation from the circulating gas stream and removal of acid gasses in an amine system. Partially cooled reactor effluent gases 516

5   (116) from the in-out exchanger 115 are further cooled in product condenser 517 (117) and conveyed through line 518 (118) to separator 519 (119) where condensed liquids are removed as product raw crude 555 (155). Overhead gases are conveyed through line 520 (120) to an amine absorber 5A (121) where acid gasses H_2S, CO_2 and SO_2 are absorbed by a counter current circulating amine solution. The recycle gases 5B flow from the top of the absorber 5A to recycle

10   hydrogen stream 129.

The rich amine solution 5C exits the bottom of the absorber, flows through an amine exchanger 5D where it is heated by exchange with hot stream amine solution 5L and enters the top of an amine stripper 5F. Absorbed acid gases are stripped from the amine solution by the application of heat to the solution in reboiler 545 (145) and are conveyed through flow line 522

15   (122) from the stripper to sulfur recovery off-site. Hot recycle gases are conveyed through line 544 (144) from the spent sand cooler 145 to provide heat for reboiler 545 (145) and the partially cooled recycled gases 5G are further cooled by cooler 5H and then flow through line 531 (131) to the suction side of compressor 132.

Lean amine solution 5J is circulated by amine circulation pumps 5K through the amine

20   exchanger 5D and amine cooler 5N to the top of the amine absorber 5A to repeat the gas cleanup process.

EXAMPLE 1

The overall mass balance for the process according to the present invention is shown in **FIG.** 6, where 1000 tons/hr of tar sand at 50°F are reacted with hydrogen to produce 665 bbl/hr

of synthetic crude oil. The following Table provides the feed and product values for processing

1000 tons/hr. of tar sand.

RAW MATERIALS		PRODUCTS
1000 TONS/HR. TAR SAND		665 BBL/HR SCO
1.6 MMSCF/HR HYDROGEN		5.2 MMSCF/HR STACK GAS
3.3 MMSCF/HR AIR		6600 LBS/HR SULFUR
0.5 MMSCF/HR NATURAL GAS		850 TONS/HR SPENT SAND

REACTOR DIMENSIONS AND MASS AND ENERGY BALANCES

5 **REACTOR 104**

Column Diameter	20.00 ft
Cross Sectional Area	314.16 ft^2
Void Fraction	0.85 (At Fluidization)
Cross Section of Sand	47.12 ft^2
10 Cross Section of Gas	267.04 ft^2
Reactor Volume	27394.26 ft^3
Bed Diameter	20.00 ft
Bed Height	87.20 ft
Time-Space Constant	0.25 hr
15 Pressure Drop	35.00 psi

TAR SAND FEED

Sand Flow Rate	1000.00 tons/hr
Density of sand	121.68 lbs./ft^3
20 Volumetric sand flow	16436.55 ft^3/hr
Sand Velocity	5.81 ft/minute
Hold-up	15.00 minutes

HYDROGEN

Hydrogen Flow Rate	238661.44 lbs/hr
25	(45226343 SCF/hr)
Cp of H$_2$	3.50 btu/lb-°F (@900°F)
Hydrogen Recycle Ratio	26.52
Hydrogen Flow Rate	45.28 SCF/hr
Hydrogen Velocity	3.02 ft/s

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OFF GAS

Gas Production	0.40 MMSCF/hr
MW	30.30 g/mole
Cp of flue gas	0.55 btu/lb-°F

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OFF GAS COMPOSITION

CO	0.30%
CO$_2$	0.20%

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H_2S	31.00%
NH_3	2.50%
C_3	66.00%

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ENERGY BALANCE

OVER-ALL CONSIDERATIONS

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Heat of Reaction	75.00 btu/lb. Bitumen
Cp Sand	0.19 btu/ton-°F
Cp Bitumen	0.34 btu/lb-°F
Cp Tarsand (sand+Bitumen)	426.70 btu/ton-°F
Sand Feed Temperature	50.00 °F

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Sand temperature	
at reactor inlet	50.00 °F
Reaction temperature	800.00 °F
Sand Feed	1,000.00 tons/hr

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TAR SAND REACTOR
REACTOR CONDITIONS

Heat required in reactor	356.03 MMbtu/hr
Heat generated in Reactor	22.50 MMbtu/hr
Additional Heat Required	335.24 MMbtu/hr
Minimum H_2 for reaction	9000.00 lbs./hr
	(1.71 MMSCF/hr)
Additional H_2 Supplied	229736.15 lbs./hr
	(43.53 MMSCF/hr)

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Total H_2 Supplied	238736.15 lbs./hr
	(45.24 MMSCF/hr)
C_1-C_2 Flow within H_2 Stream	
(at equilibrium —2%)	4594.72 lbs/hr
	(0.08 MMSCF/hr)

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Entering H_2 Temperature	1200.00 °F
Cp H_2	3.50 btu/lb-°F
Heat Supplied by C_1-C_2	1.01 MMbtu/hr
Heat Supplied by H_2	334.23 MMbtu/hr
H_2 Recycle ratio	26.53

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REACTOR BOTTOMS COOLER:
Assures Efficient Removal of Exiting Solids

Cold Hydrogen Cooler Stream	1,148.68 lbs./hr
	(0.22 MMSCF/hr)

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Heat Removed	2.73 MMbtu/hr
Entering Hydrogen Temperature	121.64 °F
Exiting Sand Temperature	791.60 °F

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SAND COOLER
SAND

Sand Flow Rate	850.00 tons/hr
Temperature of Entering Sand	791.60 °F
Temperature of Spent Sand	180.00 °F

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Cp Sand   0.19 btu/lb-°F
Heat Removed   198.59 MMbtu/hr

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HYDROGEN COOLANT FLOW
Hydrogen Flow   238736.15 lbs/hr
(45.24 MMSCF/hr)

Heat to Be Removed   182.96 MMbtu/hr
Entering Hydrogen Temperature   100.00 °F
Exiting Hydrogen Temperature   318.96 °F

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AIR COOLANT
Air Required for Combustion   250000.00 lbs/hr
(3.27 MMSCF/hr)

Cp Air   0.25 btu/lb-°F

15   Entering Air Temperature   50.00 °F
Exiting Air Temperature   300.00 °F
Heat Removed   15.63 MMbtu/hr

20   **AMINE REBOILER**
HYDROGEN SUPPLY
Entering Hydrogen Temperature   318.96 °F
Exiting Hydrogen Temperature   100.00 °F

25   AMINE BOIL-OFF
Heat Available to the system   182.96 MMbtu/hr

IN-OUT HEAT EXCHANGER
HYDROGEN TO BE HEATED
30   Hydrogen Flow   238736.15 lbs/hr
(45.24 MMSCF/hr)

Inlet H_2 Temperature   121.64 °F
Exiting H_2 Temperature   750.00 °F
Total Heat Required   525.05 MMbtu/hr

35

OFF GAS HEAT SUPPLY
Off Gas flow rate   31978.89 lbs/hr
0.40 MMSCF/hr

Condensables in vapor phase   214941.75 lbs/hr
40   MW   30.30 lb/lb-mole
Cp Vapor   0.55 btu/lb-°F
Cp Liquid   0.45 btu/lb-°F @70°F
Cp Non-Condensables   3.00 btu/lb-°F
Heat of Vaporization   65.00 btu/lb
45   Hydrogen Recycle Flow
in Stream   229736.15 lbs/hr
(*43.53 MMSCF/hr)

Inlet Temperature   800.00 °F
Exit Temperature   350.00 °F

50

PRODUCT CONDENSER/COOLER   PRODUCT SIDE
Entering Temperature   350.00 °F
Exiting Temperature   100.00 °F
Condensate   214941.75 lbs/hr

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(665.29 bbl/hr)

Heat Removal   H₂   201.02 MMbtu/hr
              Off Gas   4.40 MMbtu/hr
              Condensate 38.15 MMbtu/hr
5   Total   243.57 MMbtu/hr

COOLER REQUIREMENT   243.57 MMbtu/hr

COMPRESSOR   HYDROGEN SIDE
10 Flow Rate   755412.69 SCF/min
                                  (45.32 MMSCF/hr)

    Pressure Out   670.00 psi
    Pressure In   450.00 psi
    DP   220.00 psi
15 gamma (Cp/Cv)   1.40
    # Stages   3
    Temperature Inlet   100.00 °F
    Mechanical Efficiency 0.80 *100%
    Pb/Pa   1.14
20 Power Requirement per Stage   6366.67 hp
    Total Power Required 19100.00 hp
    Outlet Temperature   121.64 °F

    STEAM SUPPLY
25 Pressure   1500.00 psi
    Temperature   800.00 °F
    Degree Superheat   200.00 °F
    Saturation Temperature 596.20 °F
    Steam Heat Value   1364.00 btu/lb
30 Flow Rate   10894.28 lbs/hr

FIRED HEATER   PRODUCTS TO BE HEATED
    Hydrogen Flowrate   238736.15 lbs/hr
                                  (45.24 MMSCF/hr)
35 Hydrogen Temperature 750.00 °F
    Water Flow Rate   10894.28 lbs/hr
    Water Temperature   75.00 °F
    Heat Duty   517.83 MMbtu/hr

40 C₃'S (FUEL PRODUCED BY THE PROCESS)
    Flow Rate   4263.85 lbs/hr
                                  (0.04 MMSCF/hr)
    Heat of Combustion   20000.00 btu/lb
    Cp   0.60 btu/lb-°F
45 Temperature in   75.00 °F
    Heat Supplied
    (After temperature correction) 79.84 MMbtu/hr

    MAKE-UP METHANE
50 Combustion Temperature   2200.00 °F
    Heat Remaining to
          be supplied by Methane   437.99 MMbtu/hr
    Flow Rate   21653.89 lbs/hr
                                  (0.51 MMSCF/hr)

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Heat of Combustion
(After temperature correction)   20227.00 btu/lb
Temperature in   75.00 °F

COMBUSTION AIR
Air Required for Combustion   200000.00 lbs/hr
                                          (2.61 MMSCF/hr)

Air Supplied   25% Excess   250000.00 lbs/hr
                                          (3.27 MMSCF/hr)

COMPRESSOR SUCTION COOLER (5H)
OUTFLOWS
Hydrogen
                Flowrate   200000.00 lbs/hr
                Temperature   100.00 °F

Required Coolant Supply   22.42 MMbtu/hr

MATERIAL BALANCE

TAR SAND REACTOR (104)
IN FLOWS
Sand
                Flowrate   1000.00 tons/hr
                Temperature   50.00 °F
                Pressure   14.70 psia (Force Fed)

Hydrogen
                Flowrate   45.23 MMSCF/hr
                Temperature   1200.00 °F
                Pressure   635.00 psi

C_1-C_2's
                Flowrate   0.08 MMSCF/hr
                Temperature   1200.00 °F
                Pressure   635.00 psi

OUT FLOWS
Sand
                Flowrate   850.00 tons/hr
                Temperature   190.00 °F
                Pressure   600.00 psi

Off Gas
                Flowrate   43.92 MMSCF/hr
                Temperature   800.00 °F
                Pressure   600.00 psi
                Composition   wt%
                H_2 81.98
                CO   0.05
                CO_2 0.04
                H_2S 5.60
                NH_3 0.45
                C_3 11.92

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Product

5	Flowrate (Vapor Phase)	214937.52 lbs./hr
	Temperature	800.00 °F
	Pressure	600.00 psi

SAND COOLER (106, 108)   IN FLOWS

Sand

10	Flowrate	850.00 tons/hr
	Temperature	791.92 °F
	Pressure	600.00 psi

Hydrogen

15	Flowrate	45.23 MMSCF/hr
	Temperature	100.00 °F
	Pressure	500.00 psi

Air

20	Flowrate	3.27 MMSCF/hr
	Temperature	50.00 °F
	Pressure	30.00 psi

OUT FLOWS

Sand

25	Flowrate	850.00 tons/hr
	Temperature	200.00 °F
	Pressure	480.00 psi

Hydrogen

30	Flowrate	45.23 MMSCF/hr
	Temperature	313.94 °F
	Pressure	480.00 psi

Air

35	Flowrate	3.27 MMSCF/hr
	Temperature	300.00 °F
	Pressure	20.00 psi

40   **IN-OUT HEAT EXCHANGER (115)**   IN FLOWS

Hydrogen

	Flowrate	45.23 MMSCF/hr
	Temperature	147.60 °F
	Pressure	670.00 psi

45

Off Gas

	Flowrate	43.92 MMSCF/hr
	Temperature	800.00 °F
	Pressure	600.00 psi
50	Composition	wt%
	H_2	81.98
	CO	0.05
	CO_2	0.04
	H_2S	5.60

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| | | NH$_3$ | 0.45 |
| | | C$_3$ | 11.92 |

Product

5

Flowrate
(Vapor Phase)   214937.52 lbs./hr
Temperature   800.00 °F
Pressure   600.00 psi

10   OUT FLOWS
Hydrogen   Flowrate   45.23 MMSCF/hr
            Temperature   750.00 °F
            Pressure   650.00 psi

15   Off Gas   Flowrate   43.92 MMSCF/hr
            Temperature   368.63 °F
            Pressure   580.00 psi

Off Gas Composition as Above

20
Product
            Flowrate
            (Vapor Phase)   214937.52 lbs./hr
            Temperature   368.63 °F
25   Pressure   580.00 psi

PRODUCT CONDENSER/COOLER (117)   IN FLOWS

Off Gas
30   Flowrate   43.92 MMSCF/hr
            Temperature   368.63 °F
            Pressure   580.00 psi

Off Gas Composition as Above

35
Product
            Flowrate
            (Vapor Phase)   214937.52 lbs./hr
            Temperature   368.63 °F
40   Pressure   550.00 psi

OUT FLOWS
Off Gas
            Flowrate   43.92 MMSCF/hr
45   Temperature   100.00 °F
            Pressure   540.00 psi

Off Gas Composition as Above

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Product
Flowrate
(as condensate)   214937.52 lbs./hr
Temperature   100.00 °F
5   Pressure   540.00 psi

AMINE SYSTEM (121, FIG. 5)   IN FLOWS
Hydrogen
Flowrate   45.23 MMSCF/hr
10   Temperature   318.00 °F
Pressure   470.00 psi

OUT FLOWS
Hydrogen
15   Flowrate   45.23 MMSCF/hr
Temperature   100.00 °F
Pressure   450.00 psi

20   **EXAMPLE 2**

FIG. 7 shows another embodiment of the present invention. In this embodiment, a tar

sand feed is converted into a synthetic crude oil. Run of mine tar sand from trucks is dumped into

25   receiving, screening, and sizing equipment 702 for classifying tar sand at ambient temperature.

The tar sand comprises bitumen and sand. The tar sand is crushed into relatively large fluidizable

pieces that are capable of passing through a one inch mesh, or that are about one inch or less in

size. In this embodiment, crushing the tar sand into fines or pieces less than sand size is

preferably avoided to facilitate fines removal from the product stream. Limiting the amount of

30   crushing can also reduce heat generation that can adversely affect tar sand processing. Limiting

crushing can also help to preserve a water film that surrounds tar sand pieces. Tar sand pieces

typically comprise an agglomeration of sand particles, each sand particle surrounded by a film of

water and an outer layer of bitumen. On contacting a hot fluidizing flow of hydrogen during later

reaction steps, the water film can rapidly evaporate assisting the tar sand pieces to disintegrate

35   into a finely fluidized dispersion of sand particles and bitumen in hydrogen.

The crushed tar sand is conveyed through conduit 703 to feed lock hoppers 704 as the

feed for fluidized bed rector 705. The feed flow through conduit 703 and between feed lock

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hoppers 704 is controlled by pressure feeder rotary valves ("rotary valves") 703A. The bitumen

in the tar sand can contain heavy metals, such as nickel, which may catalytically promote

endothermic and exothermic reactions in reactor 705. However, supplemental catalyst such as,

for example, nickel, cobalt, molybdenum, and vanadium can be added through catalyst feed

5   conduit 704A to one of the feed lock hoppers 704 to assist catalysis provided by the heavy metals

in the mined tar sand or shale. The reactor 705 and related equipment are shown in more detail in

FIG. 8.

Recycle hydrogen in conduit 725 and fresh make-up hydrogen in conduit 725A are

conveyed to compressor 732. A first mixture of recycle hydrogen and makeup hydrogen exits

10   compressor 732 in line 733, is cooled by heat exchanger 754, and passes through line 757 to feed

lock hoppers 704. This cooled first hydrogen mixture helps to prevent the tar sand from

gumming by keeping the tar sand cool and forces the crushed tar sand into the reactor 705 which

operates at a pressure of about 600 psi. Preferably, the first hydrogen mixture reaches the lock

hoppers 704 at a temperature of about 100°F or less, and maintains the tar sand at a temperature

15   of about 100°F or less. The tar sand is fed from feed lock hoppers 704 through conduit 704B and

into reactor 705 through a feed inlet 705H, assisted by the first hydrogen mixture at 670 psi

pressure in line 757. There are three feed lock hoppers in this embodiment, but the number may

vary in other embodiments. The tar sand can be fed into the reactor approximately horizontally,

near the bottom of the reactor, and just above ceramic grid 705C. Equipment for treating mined

20   tar sand or shale feed material and for feeding the material into the reactor, such as the equipment

described above, can be referred to as a feed introducing system. Equipment for feeding tar sand

or shale feed material into the reactor, such as the feed lock hoppers 704 , conduit 703 and rotary

valves 703A, can be referred to as a feeder device.

On entering the reactor 705, the tar sand is contacted and heated by a second hydrogen

25   mixture. The second hydrogen mixture flows from fired heater 735 and into a gas inlet 705I at

24

the bottom of the reactor 705B through ceramic lined conduit line 736 at a temperature of about

1500°F and about 635 psi pressure. The second hydrogen mixture passes through a slotted fire

brick or ceramic grid 705C before contacting the entering tar sand. The flow rate and velocity of

the second hydrogen mixture are sufficient to fluidize the tar sand and to heat the tar sand to a

5   desired reaction temperature. The heated tar sand and the second hydrogen mixture react in the

reactor 705 in a fluidized bed 705E at the desired reaction temperature of about 900°F to about

1000°F, and at a pressure of about 600 psi. The second hydrogen mixture flow rate typically

exceeds the minimum needed for complete tar sand reaction with hydrogen by a factor of about

15 to about 26, and preferably by a factor of about 21. Adjustment of the second hydrogen

10   mixture flow rate may require adjustment of other reaction parameters to maintain the fluidized

bed 705E at desired pressures and temperatures. The tar sand reacts with the hydrogen mixture in

the fluidized bed 705E by endothermic hydrocracking and exothermic hydrogenating reactions.

Reaction products include substantially sulfur-free hydrocarbons that are condensable into

hydrocarbon liquids at standard temperature and pressure.

15   Reaction products including synthetic crude oil and unreacted hydrogen mixture exit the

reactor 705 through a product stream outlet 705F as an overhead or product stream through

cyclone separators 705A and into exit conduit line 710. Solids entrained in the overhead product

stream, such as sand particles and fines, are trapped by the cyclone separators 705A and are

deposited near the ceramic screen 705C at the bottom of the reactor 705B, where they are again

20   entrained in the fluidized bed 705E. Eventually, the spent sand and solids exit the reactor 705

through a conduit line 705D. The overhead stream flows through a hydrogen recycling system

wherein hydrogen is removed from the remainder of the overhead stream, treated, and returned to

the reactor.

It is advantageous to conduct the endothermic hydrocracking and exothermic

25   hydrogenation reactions in a predominantly hydrogen gas environment. The first and second

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hydrogen mixtures are mixtures of fresh make-up hydrogen and recycle hydrogen which are fed

to a compressor 732 via conduit lines 725A and 725 respectively. The recycle hydrogen contains

hydrogen and up to 5 mole percent of combined methane and ethane. The amount of combined

methane and ethane in the recycle hydrogen is maintained by a purge in a hydrogen recycle

5   system connected to the reactor 705. The volume of recycle hydrogen to fresh make-up hydrogen

is preferably about 21:1, but can vary from about 15:1 to about 26:1.

The reactor 705 is operated so as to highly agitate the reactants and ensure rapid and

complete reaction between the bitumen components and hydrogen in the reactor 705. The

residence or reaction time of the tar sand in reactor 705 is about 10 minutes, but can be between 5

10   and 20 minutes, depending on the throughput and efficiency of the reactor process. The pressure

drop from the bottom to the top of the fluidized bed 705E is about 35 psi.

Spent sand, at a temperature of about 950°F, overflows from reactor 705 into conduit line

705D through a spent solids outlet 705G. The height of the conduit line 705D may establish the

maximum height of the fluidized bed 705E. The sand then flows through spent sand lock hoppers

15   706, through conduit line 707 and into rotary coolers 708 which cool the sand from a temperature

of 950°F to about 665°F. The cooled sand can be discharged and used, for example, for land

reclamation.

The rotary coolers 708 can use ambient air fed through air intake 778 to cool the spent

sand. The air exits the rotary coolers 708 through line 779 at a temperature of about 625°F and

20   passes through a cyclone 780 to remove entrained fines. The fines are discharged through

conduit line 785. The cooling air is preheated by the spent sand, then passes to the fired heater

735 via blower 782 and conduit lines 781 and 783 where the air is used as preheated combustion

air.

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The number of feed lock hoppers 704 and spent sand lock hoppers 706 is controlled by the size of the reactor, thus more or less than the three feed lock hoppers 704 and more or less than three spent sand lock hoppers can be used in the present invention.

The reactor overhead stream from the cyclone separator 705A is discharged into line 710,

5   and then to hot gas clean-up 711. The overhead stream in line 710 exits the reactor 705 at about 950°F and enters the hot gas clean-up 711. Ceramic bag collectors or filters in the hot gas clean-up 711 remove and collect fines remaining in the overhead stream. The filters are periodically pulsated by a back flow of a 650 psi, 875°F hydrogen mixture taken from in-out heat exchanger 715 via conduit line 734A. Collected fine and solids are removed from the bottom of the hot gas

10 clean-up 711 and are collected in hot gas clean-up lock hoppers 712. The fines can be combined with spent sand and used for land reclamation. The disposal of the dry sand and fines resulting from this invention is environmentally preferable to existing wet disposal systems.

The substantially solids-free overhead stream flows from the hot gas clean-up through line 713 to the in-out heat exchanger 715. The in-out heat exchanger 715 is an indirect heat

15 exchanger wherein heat is transferred from the overhead stream to a portion of the hydrogen mixture exiting compressor 732 via conduit 733. The heated hydrogen mixture is conveyed via a conduit line 734 to the fired heater 735. The cooled overhead stream exits the in-out heat exchanger 715 through line 716.

The overhead stream in line 716 enters condenser 717 where condensable vapors and

20 gases are condensed. The overhead stream exits the condenser 717 in line 718 at a temperature of about 100°F and passes to a first separator 719 where sour water is purged from the overhead stream via line 786. The overhead stream, now purged of sour water, passes to a second separator 721 via conduit line 720 where a small vapor letdown stream is separated from the overhead stream and flows through line 722 to fired heater 735. Also, carbon compounds C_3 and above are

25 condensed and removed from the separator 721 through flow line 790 as a light substantially

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sulfur-free synthetic crude oil product stream comprising a mixture of naphtha and gas oils having an A.P.I. gravity of approximately 33.5. The crude oil product stream in conduit line 790 flows to storage and shipping. The remaining fluid in the separator 721, including recycle hydrogen, is at a temperature of about 100°F and 480 psi pressure and discharges from the

5   separator 721 as a stream in line 723 to a scrubbing system. The scrubbing system typically comprises at least one amine absorption column 724 where sulfur components, for example, hydrogen sulfide and sulfur dioxide gases, are absorbed and discharged through line 744 from regenerator 743. A sulfur recovery system can be used to recover sulfur from the sulfur components.

10   The absorber 724 can comprise, for example, a counter current circulating ethanol amine solution in intimate contact with the remaining overhead stream. The remaining fluid stream can comprise gases such as, for example, H_2S, CO_2, SO_2, NH_3, recycle hydrogen, and C_1 and C_2 hydrocarbons. H_2S, CO_2, SO_2, and NH_3 are removed from the remaining fluid stream by the absorber 724. Remaining hydrogen, C_1 and C_2 hydrocarbons form the recycle hydrogen mixture

15   and flow through line 725 to compressor 732.

The rich amine solution having absorbed H_2S, CO_2, SO_2 and NH_3 is discharged from the absorber 724 through line 740 and flows through an amine heat exchanger 741. In the amine heat exchanger 741 the rich amine solution is heated by exchange with hot amine solution in line 750 which is returning from amine regenerator 743 to the absorber 724. The heated rich amine

20   solution flows through line 742 and enters the top of the amine regenerator 743. Absorbed acid gases are stripped from the rich amine solution by further heating the rich solution using steam from a steam reboiler 745 Heat for the reboiler 745 is supplied by steam from the fired heater 735 steam recovery system.

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Lean amine solution is discharged from the regenerator 743 in line 748 and is circulated by an amine circulation pump 749 through amine exchanger 741 and amine cooler 752 to the top of the amine absorber 724.

Recycle hydrogen, and C_1 and C_2 hydrocarbons flow through line 725 to compressor 732 and are mixed with make up fresh hydrogen in line 725A at a pressure of 450 psi and a temperature of about 100°F. The recycle gas stream is also at a pressure of 450 psi, which is the lowest pressure in the system. The compressor 732 is driven by a high pressure steam turbine 763. High pressure steam supply in line 762 comes from the fired heater steam system at 900 to 1500 psi and a temperature of 800°F, which is super heated by 200°F in the fired heater 735. Exhaust steam in line 764 is condensed in condenser 765 and along with make up water is fed to the fired heater 735 for preheating and reuse as boiler feed water make up.

The compressor 732 pressurizes the recycle hydrogen mixture and make-up hydrogen from 450 psi to approximately 670 psi and 187 °F and discharge the hydrogen mixture into line 733. A portion of the hydrogen mixture in line 733 is the first hydrogen mixture and is delivered to heat exchanger 754 via line 733A. Another portion of the hydrogen mixture in line 733 is the second hydrogen mixture and is delivered to the in-out heat exchanger 715.

The heat exchanger 754 cools the first hydrogen mixture from about 187°F to 100°F. A portion of the first hydrogen mixture in line 757 flows into line 756 and to a C_1 and C_2 hydrocarbon pressure swing adsorption ("PSA") system 755. The PSA system helps to maintain the C_1 and C_2 hydrocarbon level in the first and the second hydrocarbon mixture at about 2% - 3%. C_1 and C_2 hydrocarbon purged from the first hydrocarbon mixture is discharged through line 758 and combined with the gas in line 22 which is delivered to the fired heater 735. Purified hydrogen produced by the PSA 755 flows through line 756A and back to the suction of compressor 732 via line 725.

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The second hydrogen mixture is preheated to 875°F in the in-out heat exchanger 715 by the overhead stream at 935°F. Preheated air conveyed through feed line 783 is combusted with fuel in the fired heater 735 and elevates the temperature of the second hydrogen mixture that is conveyed through line 734 from in-out heat exchanger 715. The fuel that is combusted is

5   obtained from the natural gas line 759 and purge gas line 722. The hydrogen mixture circulates through the fired heater 735 and exits through line 736. The second hydrogen mixture provides the heat required to maintain reaction in the reactor 705.

Waste heat from the radiant section of direct fired heater 735 is recovered in convection sections 735A, 735B and 735C. Steam and water are discharged from a steam drum 760 into the

10   fired heater 735. Heated steam is returned to the drum via line 773. Steam separated from water in drum 760 is discharged into line 761 and introduced into convection section 735A where the steam temperature is raised from about 596°F to about 800°F. After passing through convection section 735A, the super heated, high pressure steam is conveyed through line 762 to drive the steam turbine 763. Reduced temperature and pressure steam from turbine 763 is conveyed to

15   steam condenser 765 and the condensate recirculated via line 767. The flow from pump 766A is conveyed through line 767 and combined with make-up water. The water being conveyed in line 767 is introduced into convection section 735C, heated and discharged through line 736 for further processing, such as aeration.

The following table shows material flows and operating conditions an operating reactor

20   system.

FLUIDIZED BED REACTOR:

Reactor (fluidized bed) Temperature	950	°F
Reactor (fluidized bed) Pressure	600	psi
H_2 Recycle Ratio	21.09	
Catalyst Flow Rate into Reactor	1255.07	lbs/hr
Tar Sand Flow into Reactor	2520	tons/hr
Tar Sand Feed Inlet Temperature	50	°F
Hydrogen Mixture Flow Rate into Reactor	60.4	MMSCF/hr
Hydrogen Mixture Gas Inlet Temperature	1500	°F

ROTARY COOLERS:

Air Temperature at Intake	50	°F
Air Temperature (exiting)	623	°F
Sand Entering Temperature	950	°F
Sand Exiting Temperature	665	°F

IN-OUT HEAT EXCHANGER:

Overhead Stream Entering Volumetric Flow Rate	58.31	MMSCF/hr
Overhead Stream Entering Temperature	950	°F
Overhead Stream Exiting Temperature	516	°F
Hydrogen Mixture Entering Flow Rate	60.4	MMSCF/hr
Hydrogen Mixture Entering Temperature	185	°F
Hydrogen Mixture Exiting Temperature	875	°F

FIRED HEATER:

Fuel Consumption (natural gas eqivalent)	1.2	MMSCF/hr
Vapor Let-Down & PSA Off Gas Fuel Supply (natural gas equivalent)	0.56	MMSCF/hr
Make-up Fuel Supply	0.64	MMSCF/hr
Combustion Air Entering Flow Rate (@65% excess)	26.59	MMSCF/hr
Steam Production Rate (@ 1500 psi)	228,996	lbs/hr
Hydrogen Mixture Entering Flow Rate	60.4	MMSCF/hr
Hydrogen Mixture Entering Temperature	875	°F
Hydrogen Mixture Exiting Temperature	1500	°F

COMPRESSOR:

Power Required From Turbine	40,148	h.p.
Steam Flow Rate to Turbine	228,996.1	lbs/hr
Steam Pressure Entering Turbine	1500	psi
Steam Temperature Entering Turbine (200°F superheat)	800	°F
Hydrogen Mixture Entering Flow Rate	60.7	MMSCF/hr
Hydrogen Mixture Compressor Entering Temperature	100	°F
Hydrogen Mixture Compressor Entering Pressure	450	psi
Hydrogen Mixture Compressor Exiting Temperature	185	°F
Hydrogen Mixture Compressor Exiting Pressure	670	psi

PRODUCT CONDENSER/SEPARATOR:

Product Fluid Stream Entering Flow Rate	58.3	MMSCF/hr
Product Fluid Stream Entering Temperature	516	°F
Recycle Hydrogen Mixture Exiting Temperature	100	°F
Synthetic Crude Oil Flow Rate	1255	bbl/hr

AMINE SYSTEM:

Amine Recirculation Flow Rate	50,400	lbs/hr
Ammonia Production	1478	lbs/hr
Elemental Sulfur Production	17260	lbs/hr

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While particular embodiments of the present invention have been illustrated and described herein, the present invention is not limited to such illustrations and descriptions. It is apparent that changes and modifications may be incorporated and embodied as part of the present invention within the scope of the following claims.